Refereed Proceedings

The 12th International Conference on Fluidization - New Horizons in Fluidization Engineering

Engineering Conferences International *Year* 2007

Hydrodynamics of Gas-Solids Bubbling Fluidized Beds Using Polyethylene Resin

Bangyou Wu* John Shepperson† Loni van der Lee‡
Céline Bellehumeur** Apostolos Kantzas††

*University of Calgary, bangyouwu@hotmail.com
†University of Calgary
‡University of Calgary
**University of Calgary
††University of Calgary, akantzas@ucalgary.ca

HYDRODYNAMICS OF GAS-SOLIDS BUBBLING FLUIDIZED BEDS USING POLYETHYLENE RESIN

Bangyou Wu[1,2], John Shepperson[1,2], Loni van der Lee[1,2], Céline Bellehumeur[2], Apostolos Kantzas[1,2]
[1]Tomographic Imaging and Porous Media Laboratory
[2]Department of Chemical and Petroleum Engineering
Schulich School of Engineering
University of Calgary, Calgary, AB, Canada T2N 1N4
T: +1-403-220-8907; F: +1-403-282-5060; E: akantzas@ucalgary.ca

ABSTRACT

Pressure fluctuations were measured in three gas-solids bubbling fluidized beds (10cm, 20cm, and 30cm diameter, 1.0m high) using polyethylene particles of different sizes and particle size distributions. Both bed scale and particle size distribution significantly affect the gas-solids flow behavior. Bubble diameters measured from X-ray fluoroscopy in the 10cm diameter column were compared to various correlations in the literature.

INTRODUCTION

Hydrodynamics of gas-solids fluidized bed have been investigated extensively, however, the flow behavior of porous particle like polyethylene is still not well understood. Bubble properties, such as bubble diameter, play an important role in the hydrodynamics of gas-solids flow. In the literature, various literature correlations [1-5] were proposed to calculate bubble properties, such as bubble diameter and bubble velocity. It is necessary to verify which correlation is most suitable for porous particles. In this study, X-ray fluoroscopy, a non-intrusive imaging technique, was used to characterize the bubble properties in the 10 cm diameter fluidized bed. Estimated bubble diameters were compared to published correlations to find the most closely matched ones.

Scale-up is the eventual purpose of research work on fluidized beds. In this work, similar experiments were carried out in columns of different diameters. Also, particle size distribution effect was moreover investigated in the 10 cm column. As pressure is the most convenient measurement for an industrial fluidized bed reactor, pressure fluctuations were measured for the scale and particle size effects on hydrodynamics.

EXPERIMENTAL

The experiments were conducted in three cold-model Plexiglas fluidized bed reactors designed based on the dimensions of the UNIPOL reactor. Each column had a windbox filled with marbles or plastic sphere to disperse the gas, a porous plate distributor, a 1-m bed section and an expansion section of 1.5 times of the

The 12th International Conference on Fluidization - New Horizons in Fluidization Engineering, Art. 125 [2007]

column diameter. Figure 1 shows a picture of the 3 columns used in this study. Air was used as fluidization gas. The air was compressed and the flow rate was controlled using pressure regulator and rotameters. A cyclone was used to remove the fines before the air was vented to the atmosphere.

Figure 1. Experimental setup of fluidized bed columns

Two linear-low-density-polyethylene resins were used as the solids phase. Resins were collected directly from industrial reactors and supplied in powder form by Nova Chemicals. The first was a butene-ethylene copolymer (P1) with particle size distribution of 500-700 μm and a plaque density of 918 kg/m^3. The other was a hexene-ethylene copolymer (P2) with wide size distribution (165-1500 μm), a mean particle size distribution of 850 μm and a plaque density of 918 kg/m^3. These particles, which are irregular and porous, fall in Geldart B group.

The minimum fluidization velocity (U_{mf}) was determined by measuring the bed pressure at different velocities. Transducers were calibrated to establish the relationship of pressure versus voltage prior to the experimental measurements. Table 1 lists U_{mf} for all particles tested for pressure fluctuation measurement.

Four pressure transducers (Schlumberger Solartron, model 8000 DPD) were connected to column wall pressure ports located at a height h=0, 17, 34, and 53 cm above the distributor using 0.4 cm plastic tubes. Each transducer measured 0-200 inches water (0-49.8 kPa). Tubing length ranged from 50cm to 90cm. An A/D converter, a 12-bit PC-LPM-16 card from National Instruments, and a personal computer were used for data acquisition. Digital pressure signals were stored into the computer through a Labview™ program. The pressure taps were mounted flush with the column wall and the inside of the pressure taps was covered with a wire mesh to prevent solids from entering the tubing or transducer. The pressure data was collected at a rate of 500 Hz for 60 seconds at each flow rate. Each operating condition was sampled 20 times. More detail about the experimental measurement can be found elsewhere (6).

Table 1. Minimum fluidization velocity under various experimental conditions

Column diameter (cm)	Particle size distribution (μm) (*mean size*)	Static bed height (cm)	Resin	U_{mf} (cm/s)
10	< 425 (*212*)	40	P2	3.9
10	425-707 (*566*)	40	P2	9.5
10	707-1000 (*853*)	40	P2	20.3
10	165-1500 (*850*)	40	P2	7.7
10	500-700 (*600*)	60	P1	6.3
20	500-700 (*600*)	60	P1	6.3
30	500-700 (*600*)	37*	P1	6.3

* The static bed height is only 37 cm due to the limited amount of particles available.

The X-ray fluoroscopy was used to measure the real time gas-solids flow behavior in the 10 cm column. The minimum fluidization velocity of a sieved fraction of P2 (165-1500 μm with mean size of 850 μm) is 7.7 cm/s. Three superficial gas velocities (2.5, 3.0, 3.5 U_{mf}) were tested. The captured images were processed using image analysis software developed in-house to identify the bubble boundaries. Images were enhanced and then segmented. Local thresholding identified each bubble and then examined the intensity histogram to determine bubble boundaries. Once the boundaries of the blobs were identified the bubble diameter was estimated.

RESULTS AND DISCUSSION

Bubble Diameter

The experimental values for bubble diameter determined from X-ray fluoroscopy in the 10 cm column were compared with values predicted by a number of published correlations [1-5]. Figure 2 shows the comparison between model predictions and one set of experimental results (U_g=3U_{mf} selected as an example). At high superficial velocities ($U_g \geq 3U_{mf}$) the correlation of Cai et al. (1) matches with experimental results very well. At lower velocities, however, the correlations overpredict the bubble diameter. This is likely due to the experimental limitation for measuring small bubbles at low superficial gas velocity. X-ray fluoroscopy measurements only capture the vertical 2-dimensional image of the fluidized bed. If there are many small-diameter bubbles at the same cross-section of the column, it is very hard to detect all the bubbles due to the averaging effect on the overall density. Therefore, the measured bubble diameters may be smaller than those predicted from correlations.

Among the correlations that match reasonably well with experimental results, it is found that they either are known to work best with Geldart group B particles or alternatively, have been developed using data obtained from columns with small diameters (≤10cm). To test if the first case is true, the predicted results from those correlations were compared with experimental results from two new particle size distributions of the same polyethylene powder, with mean particle sizes of 212μm (0-425μm) and 566μm (425-707μm) respectively (Figures 3 and 4).

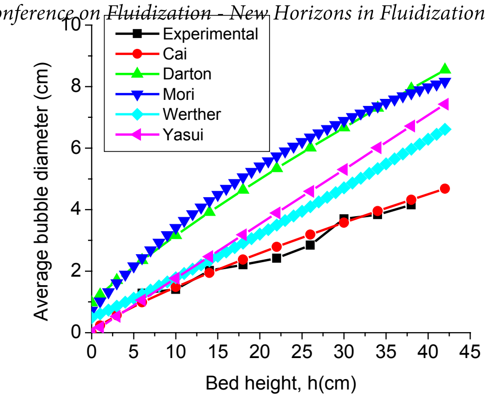

Figure 2. Comparison of bubble diameter from experimental and prediction from correlations at $3.0U_{mf}$=23.1 cm/s (U_g-U_{mf}=15.40cm)

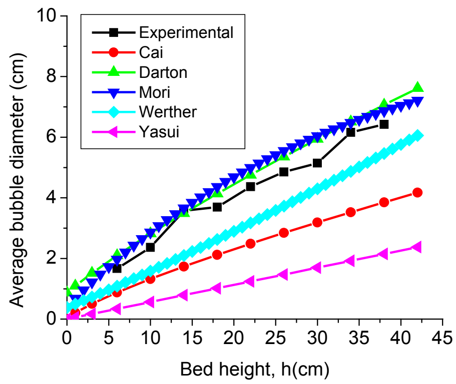

Figure 3. Comparison of bubble diameter from experimental and prediction from correlations at 15.45 cm/s (U_g-U_{mf} = 11.55 cm/s) for mean particle sizes of 212 μm

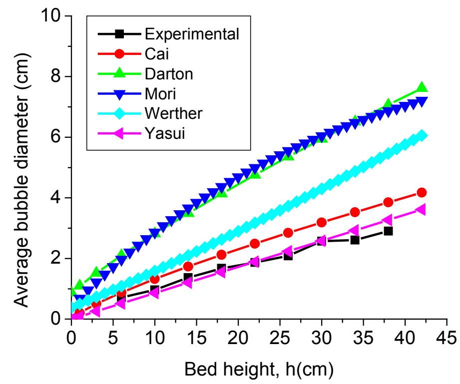

Figure 4. Comparison of bubble diameter from experimental and prediction from correlations at 21.05 cm/s (U_g-U_{mf} = 11.55 cm/s) for mean particle sizes of 566 μm

Results obtained with the narrow size distribution with a mean of 566 μm (Figure 4) are similar to those obtained with the broader particle size distribution (165-1500 μm) at similar U_g (Figure 2). This would be anticipated because the mean particles size is large. The bubble diameter profile changes with lower mean particle diameter as shown in Figure 3. The correlations of Yasui and Johanson, Cai et al. and Werther et al. underestimate the values of the bubble diameter, while the Darton et al (4) and Mori and Wen (5) correlations provide a better match with the experimental results. This shows consistency with the original papers presented by these authors. Mori and Wen found that their predictions showed excellent agreement with experimental results for sand with mean particle sizes of 128 μm and 150μm (Geldart B). In the work carried out by Darton et al., it was also found that there was excellent agreement between the model and experimental data from quartz sand, with a mean particle size of 125 μm (Geldart B). Thus, it would appear that from results obtained so far, the Darton and Mori and Wen correlations are best suited to relatively small diameter particles. The influence of the column diameter on the bubble diameter profile is another issue that will be further addressed in future research work.

Bed Scale Effect

Bed scale effect on the hydrodynamics was shown in Figure 5 and 6. As pressure fluctuation mainly reflects the global flow behavior, measurement position does not significantly affect the dynamic behavior. Only results from one measurement position at *h*=34cm for all the three columns is shown. More detail of calculation of the parameters, namely average cycle time (ACT) and Kolmogorov entropy (*K*), and wavelet analysis can be found elsewhere (7). The software used for calculating ACT and *K* was RRCHAOS developed by Schouten and van den Bleek (8). The minimum fluidization velocity for P1 particle was found to be 6.3 cm/s for all the three columns used in this work (10, 20 and 30 cm diameter). For $0<U_g<6.3$ cm/s, pressure signals were dominated by electrical noise. Therefore, the estimated parameters showed similar very low ACT and very high *K* (around 150 bits/s). With an increase of U_g, ($6.3<U_g<15$ cm/s), ACT increased by a factor of 2 (minimum) while *K* showed a significant decrease. Large-amplitude and low-frequency cyclic behavior of pressure fluctuation becomes dominant with the formation of bubbles above U_{mf}. This relatively regular behavior caused the dramatic decrease of *K* (Figure 5). As the bubbles were still relatively small, the scale effect was not significant with the increase of U_g within this intermediate velocity range. For higher superficial gas velocities (U_g>15 cm/s), the parameters show the difference due to different diameter for the three columns. ACT keeps increasing with an increase of U_g, while *K* decreases slowly with an increase of U_g. The large ACT for 10 cm column is likely due to the wall effect and bubble coalescence. Visual observation of the bed behavior suggested that slugging might have commenced at high U_g, as large bubbles were seen close to the column wall. The calculated slugging onset velocity by Baeyens and Geldart (9) is around 17.2 cm/s, which is very consistent to our experimental results. Certainly a lot of experimental results in the 10 cm column were operated in the slugging regime. Slugging behavior caused very large ACT. Consistently, high ACT corresponds to low *K*, which is indicative of a less chaotic and more regular flow behavior. However, for the operation of a reactor, small ACT and high *K* may be favorable for enhanced heat and mass transfer. Therefore, the scale of the fluidized bed affects significantly the hydrodynamic behavior and should be considered in the scale-up process and fluidized bed modeling/simulation. Also, further comparison of our experimental results with published work is needed.

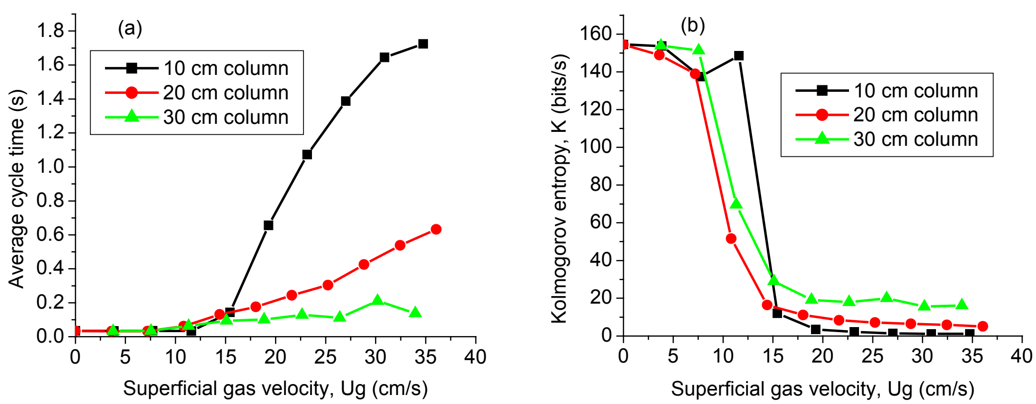

Figure 5. Bed diameter effect on ACT and *K* as a function of U_g at *h*=34 cm

Wavelet analysis was performed on the pressure fluctuation data to investigate the scale of dynamic flow in different columns. Signal series were decomposed into approximations and details of different scales, details on the mathematical treatment are presented elsewhere (10). Different scales were considered: micro-scale signals characterizing particle movement in the dense phase or in the gas phase, meso-scale signals describing medium-sized bubble/dense phase behavior or their interaction, and macro-scale accounting for large-sized bubble and the effect of the column on the system behavior. Standard deviations (SD) of detail coefficients for 6 scales are presented in Figure 6.

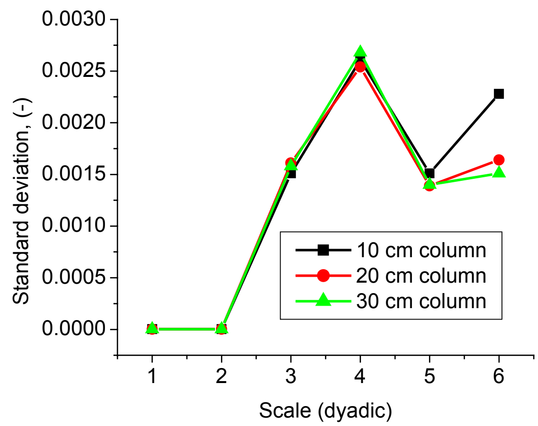

Figure 6. Bed diameter effect on standard deviation of detail wavelet coefficients at h=34 cm and U_g=3U_{mf}

The profiles shown in Figure 6 provide some information about the amplitude of pressure fluctuations (magnitude of SD). For small scale (<5), bed diameter has no notable effect on the standard deviation of wavelet coefficients. At larger scale (≥5), the standard deviation of the coefficients decreases with an increase of bed diameter. Bed diameter affects more significantly large-scale flow characteristics, more important fluctuations being detected in the small-diameter column. It is necessary to investigate larger scale (>6) to fully understand the scale effect on the detail coefficients.

Particle Size Distribution Effect

The effects of particle size and particle size distribution on the flow behavior of our gas-solid system were evaluated based on pressure fluctuation data. Results obtained at only one position (h=34 cm) are shown in Figures 7 and 8, as they are found to be representative of those collected at different bed height in the 10 cm column. As expected, smaller size particles tend to fluidize more easily. The profiles for ACT and K with respect to the superficial gas velocity are not linear. For each particle size fraction, a critical velocity can be identified based on the results presented in Figure 7 at which ACT shows a sharp increase while K rapidly drops to very small values. This phenomenon represents a transition between minimum fluidization and slugging behavior. For narrow size distribution, the transition velocity increases with an increase of mean particle diameter, as U_{mf} increases with an increase of particle size. The transition from bubbling to slugging is clearer from profiles of K in Figure 7(b). At minimum fluidization, K is very high (around 150 bits/s), indicating very chaotic pressure fluctuation due to complex flow behavior of small bubbles. Pressure fluctuations are generally small amplitude and high frequency. With the increase of U_g, mean bubble diameter also increases. Larger bubble caused more large-amplitude and low-frequency pressure fluctuations, so that K decreases and ACT increases with an increase of U_g. When slugging commences, the regular passage of large voids causes almost periodic behavior, thus ACT is very large and *K* is very low (< 20bits/s). Because *K* is more sensitive to U_g than ACT in bubbling regime, it is more likely to determine the onset velocity of slugging regime using *K*. From the profiles shown in Figure 7, the decrease of *K* for the broad size distribution (165-1500μm) is more gradual than narrow size distributions, which indicates different bubbling behavior due to the wide particle size distribution. With the presence of smaller or larger particles, bubbling behavior, thus heat and mass transfer may be enhanced.

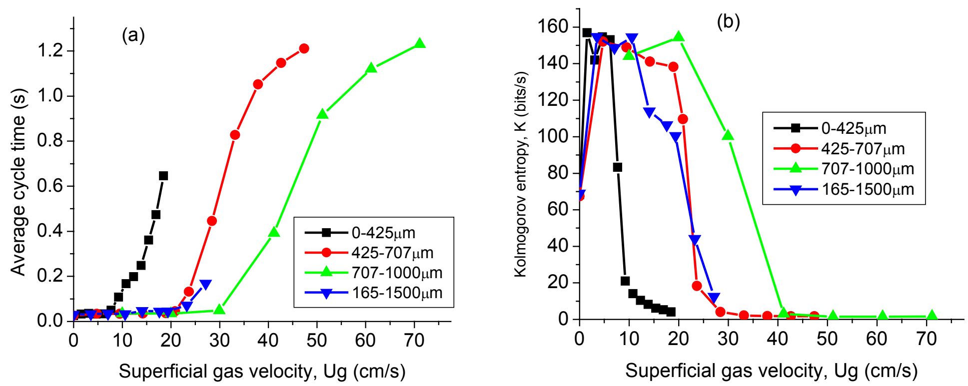

Figure 7. Particle size distribution effect on ACT and *K* as a function of U_g at *h*=34cm

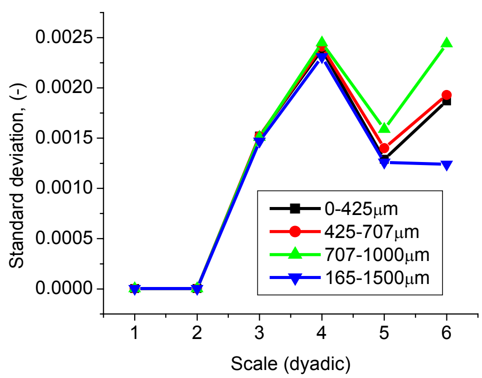

Figure 8. Particle size distribution effect on standard deviation of detail wavelet coefficients at *h*=34 cm and U_g=3U_{mf}

The powder particles with a broad size distribution (165-1500 μm) showed similar fluidization characteristics to those obtained with particles with a size fraction of 425-707 μm (Figure 7). From the wavelet analysis (Figure 8), we observed that only the large-scale behavior (scale≥4) was affected by the particle size and size distribution. Larger particles showed higher standard deviations of wavelet coefficients. Interestingly, the mixed particles (165-1500 μm) showed smallest standard deviation for all the scales.

This result suggests that wide particle size distributions have the least chance to form large bubbles under the same superficial gas velocity ratio (U_g/U_{mf}). This, however, needs to be verified using results from X-ray fluoroscopy measurements and further compared with data reported in the literature.

CONCLUSIONS

Bubble diameter determined from X-ray fluoroscopy imaging of polyethylene resins was found to agree well with some empirical correlations. Predictions obtained using the correlation proposed by Cai et al. shows good agreement with X-ray fluoroscopy measurements for relatively large particles, a category which describes most powder particles used in this work. Darton et al. and Mori and Wen correlations were found to be much closer to the experimental results for relatively small powders.

Our results clearly showed that both the bed diameter and the particle size distribution greatly affect the measured hydrodynamic behavior. This should be considered in the scale-up procedure and hydrodynamic modeling or simulation for porous particles like polyethylene resins.

ACKNOWLEDGEMENTS

Financial support from the Natural Science and Engineering Research Council of Canada and the Canada Research Chairs Program is gratefully acknowledged.

NOTATION

ACT	average cycle time [s]	SD	standard deviation [-]
h	bed height from distributor [cm]	U_g	superficial gas velocity [cm/s]
K	Kolmogorov entropy [bits/s]	U_{mf}	minimum fluidization velocity [cm/s]

REFERENCES

1. Cai, P., Schiavetti, M., De Michele, G., Grazzini, G.C., Miccio, M. Quantitative estimation of bubble size in PFBC, Powder Technology, **80** (2), 99-109, 1994.
2. Yasui, G. and Johanson, L.N., Characteristics of Gas Pockets in Fluidized Beds, AIChE J., **4**, 445-452, 1958.
3. Werther, J., Davidson, J.F., and Keairns, D.L., Influence of the Distributor Design on Bubble Characteristics in Large Diameter Fluidized Beds, in: Fluidization, Cambridge University Press, London, p.7-12, 1978.
4. Darton, R.C., La Naueza, R.D., Davidson, J.F., and Harrison, D., Bubble growth due to coalescence in fluidized beds, Trans. Inst. Chem. Eng. **55**, 274-280, 1977.
5. Mori, S. and Wen, C.Y. Estimation of bubble diameter in gaseous fluidized bed, AIChE J., **21**, 109-115, 1975.
6. van der Lee, L., Analysis on Pressure Signals from a Polyethylene Gas-Solid Fluidized Bed, M.Sc. Thesis, University of Calgary, Calgary, Canada, 2006.
7. Wu, B.Y., Kantzas, A., Bellehumeur, C., and Kryuchkov, S., in: "5th World Congress on Particle Technology", Conference proceeding CD, 2006.
8. Schouten, J.C. and van den Bleek, C.M., RRCHAOS, ver 2.5, Delft University of Technology, The Netherlands, 1994.
9. Baeyens, J. and Geldart, D. An investigation into slugging fluidized beds, Chem. Eng. Sci., **29**, 255-265, 1974.
10. Wu, B.Y., Kantzas, A., Bellehumeur, C.T., He, Z.X., and Kryuchkov, S., Multiresolution Analysis of Pressure Fluctuations in a Gas-Solids Fluidized Bed: Application to Glass Beads and Polyethylene Powder Systems, Chem. Eng. J., (Accepted, 2006).